

GASIFICATION OF CANADIAN ANTHRACITE

by

A.P. Watkinson, Principal Investigator
G. Cheng, Research Engineer
R. Quan, Research Engineer

Department of Chemical Engineering
The University of British Columbia

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SUMMARY

Mount Klappan anthracite has been gasified with oxygen-steam and air-steam mixtures in a 0.30-m dia. atmospheric pressure gasifier operated in the spouted bed mode. Coal rates of about 12-23 kg M.A.F./h were used. In some trials char was recycled from the primary cyclone to the bed. A typical gas from oxygen-steam gasification at about 915°C with oxygen/coal mass ratios around unity, contained 32% H₂, 28.5% CO, 1% CH₄, 0.5% N₂ and 38% CO₂. The heating value of this gas is about 7.8 MJ/m³. The high CO₂ level increased still further as the temperature of the gasifier was raised to over 1000°C by increasing the oxygen/coal ratio to 2.2. Under these conditions some 80% of the carbon was converted, and a gas containing 26.5% H₂, 25% CO, 0.4% CH₄ and 48% CO₂ is produced. These high CO₂ levels reflect the inefficiency of the pilot scale gasifier, and the relatively low reactivity of the anthracite. Air-steam gasification gave correspondingly lower gas heating values due to the presence of the nitrogen. Controlled ash agglomeration was not achieved in the few trials attempted. A single test with K₂CO₃ addition as a gasification catalyst showed significantly improved gas quality, efficiency and steam decomposition. There was no apparent difference between operation with high ash (30%) and low ash (10%) coal.

INDEX

	PAGE
Summary	i
List of Figures	iii
List of Tables	iv
1. Introduction	1
2. Apparatus	3
3. Gasifier Operating Procedures	5
4. Experimental Plan	7
5. Results and Discussion	
5.1 Heating Values, Carbon Conversion and Other Performance Indicators	7
5.2 Typical Experimental Result	11
5.3 Effect of Oxygen/Anthracite Ratio	11
5.4 Ash Agglomeration	14
5.5 Catalytic Run	15
5.6 Internal Temperature and Composition Profile	17
5.7 Low Ash Anthracite Run	17
5.8 Gas Composition versus Kinetics-Free Model Predictions	20
6. Conclusions	21
References	22
Figures 1 to 15	23-34
Appendix A Statement of Work	35
Appendix B Kinetics-Free Model	37

LIST OF FIGURES

NO.	TITLE	PAGE
1.	Schematic Diagram of Apparatus	23
2.	Result of Typical Experiment	24
3.	Carbon Conversion and Temperature Versus Oxygen to Anthracite Feed Rate Mass Ratio	25
4.	Gas Heating Value versus Oxygen to Anthracite Feed Rate Mass Ratio	26
5.	Gas Composition versus Oxygen to Anthracite Ratio	26
6.	Efficiency Factor versus Oxygen to Anthracite Ratio	27
7.	Efficiency Factor versus Ratio of Oxygen to Feed Plus Recycle Rate	27
8.	Carbon Conversion and Gas Heating Value versus Average Bed Temperature	28
9.	Typical Temperature Profiles during Ash Agglomeration Attempt	29
10.	Internal Gas Composition and Temperature Probe	30
11.	Radial Gas Composition Profiles at Three Different Axial Positions - Run 160	31
12.	Axial Gas Composition Profiles - Run 160	32
13.	Axial Gas Composition Profile - Run 158	33
14.	Experimental versus Predicted H_2/CO mol Ratio	34
15.	Experimental versus Predicted Dry Gas Heating Values	34

LIST OF TABLES

	PAGE
Table 1 Analyses of Anthracite Samples	2
Table 2 Dimensions of Coal Gasifier	4
Table 3 Particle Size Distribution of Anthracite Feed, Bed Material and Cyclone Product	4
Table 4 Experimental Plan	8
Table 5 Summary of Experimental Results	9
Table 6 Comparison of Maximum Gas Heating Value Gasification Conditions for Anthracite and Other Coals	12
Table 7 Catalytic versus Non-catalytic Gasification Runs	16
Table 8 Probe Positions and Gas Temperatures	18
Table 9 Gasification of Low Ash and High Ash Coals	19

1.0 Introduction

In gasification of coal it is generally found that low rank coals, particularly those which are high in moisture and volatiles are most readily gasified. During the de-volatilization stage which occurs at relatively low temperatures, a porous char is produced by the release of the volatiles. This resulting char is of high surface area, and aided by the presence of mineral matter which can catalyze the attack of steam and CO_2 on the carbon, the gasification reactivity is high. Anthracite, being of high rank, and thus low in volatiles tends to be relatively unreactive, as it does not yield a highly porous char, on which the gasification reactions may proceed. Thus it requires higher temperatures and more severe conditions to gasify. In this project a B.C. anthracite from Mt. Klappan supplied by Gulf Resources Canada Ltd. was evaluated in a 0.305 - m. dia. atmospheric pressure gasifier. The objectives of this work were to study the gasification of Mt. Klappan anthracite in order

- (i) to provide an option for utilization of high ash content Canadian anthracite via the gasification process
- (ii) to complement CANMET in-house research on the correlation of gasification reactivity with coal rank.

Appendix A lists the specific tasks required for the project.

The high-ash anthracite sample supplied was not only higher in ash (at 30%) than projected (at 20%), but also somewhat variable (Table 1). Thus the initial sampling of the lot gave the composition listed as anthracite A. A subsequent sampling for the latter 3 runs gave the analysis listed for Anthracite B. The ash levels between these two samples differed by 5.5% on a dry basis, the essential difference being made up of carbon. In the latter stages of the contract, a comparative gasification test was made using a lower-ash anthracite, labelled Anthracite-C, which contained 10.3% ash.

Table 1: Feedstocks Used in Hot Gas Cleanup Studies

Feedstock	Mt. Klappan Anthracite-A (high ash)	Mt. Klappan ^{**} Anthracite-B (high ash)	Mt. Klappan Anthracite-C (low ash)
Elemental Analysis (% Wt. Dry Basis)			
C	60.94	66.22	83.08
H	2.01	2.06	2.84
N	0.79	1.06	1.06
S	0.40	0.33	0.43
Cl	0.06	0.01	0.02
O [*]	3.38	3.42	1.83
Ash	32.42	26.89	10.74
Proximate Analysis (% Wt. As Rec'd)			
Moisture	7.16	5.48	4.37
Ash	30.11	25.42	10.27
Volatiles	7.69	6.99	6.56
Fixed Carbon	55.06	62.11	78.80

* by difference

** Used for Runs 158 - 160

2. Apparatus

The apparatus used for the main gasification study has been described previously (1,2), but will be described here for completeness. The gasifier specifications are given in Table 2, and a schematic diagram of the unit is shown as Figure 1. The reactor itself consists of three major sections; the main body is a 10 gage carbon steel column with outside diameter 0.50 m and height 1.37 m, and is refractory lined to an inside diameter of 0.30 m to withstand the high temperatures. Below the main section is a 0.46 m high 316 stainless steel cone which is also refractory lined to a 50 degree angle. The uppermost section of the reactor is a larger diameter (0.76 m) disengaging section, which reduces the amount of particulates leaving the gasifier.

During the pre-heating stages of the experiment, hot flue gas from the propane burner enters the cone bottom and flow upwards, spouting and heating a pre-determined charge of bed material.

Once the reactor has been heated to the desired temperature, anthracite is pneumatically transported from the vibra-screw feeder into the reactor via the 2.54-cm dia. solids transport line. It is also possible to feed a second solid material (such as potassium carbonate) into the solids transport line by means of the auxiliary feeder/rotary valve assembly.

During gasification, flue gas and some solids carry over from the reactor into the gas cleanup section. In the usual operation with conventional cyclones particulate material is removed by the two cyclones in series, and when necessary, dust can be recycled from the first cyclone into the reactor through a rotary valve. Char from both cyclones is collected in a char (dust) drum and weighed and analyzed following each experiment. Hot gas leaving the cyclones is cooled in an indirect heat exchange with water

Table 2: Gasifier Specifications

Gasifier:	
Column	0.50-m O.D. x 0.305-m I.D. x 1.37-m high
Cone	Angle 60°, height 0.46-m
Disengaging Section	0.76-m O.D. x 0.56-m I.D. x 0.61-m high
Spouted Bed Orifice Diameter	3.2 cm
Startup Burner	176 kW max. propane fired
Cyclone - 1st Stage: 20.6 cm dia. x 27.3 cm cylindrical height x 33.7 cm cone height	
Cyclone - 2nd Stage: 19.4 cm dia. x 26.7 cm cylindrical height x 48.3 cm cone height	
Gas Coolers (3)	Each 2.2 m length x 7.6 cm I.D. x 10 cm dia. jacket
Gas Knock Out Drum	56 cm dia. x 90 cm height

Table 3: Particle Size Distribution of Anthracite Feed, Bed Material and Cyclone Product

Feed		Bed Material*		Cyclone Product*	
Avg. Size (mm)	Weight Fraction	Avg. Size (mm)	Weight Fraction	Avg. Size (mm)	Weight Fraction
4.75	0.007	3.35	0.053	2.18	0.002
4.05	0.064	2.86	0.063	1.85	0.003
2.86	0.150	2.18	0.045	1.44	0.004
2.03	0.177	1.85	0.052	0.89	0.004
1.35	0.260	1.44	0.127	0.48	0.014
0.80	0.165	0.89	0.233	0.24	0.070
0.48	0.082	0.48	0.133	0.18	0.133
0.30	0.035	0.28	0.104	0.13	0.157
0.22	0.034	0.18	0.063	0.091	0.171
0.15	0.018	0.13	0.042	0.064	0.099
0.12	0.004	0.09	0.032	0.045	0.142
0.05	0.006	0.05	0.056	0.038	0.201

d_p (mm)	0.831	0.302	0.071
$\bar{d}$ (mm)	1.62	1.06	0.12

* Run 149 $\bar{d} = \sum x_i d_i$ $d_p = (\sum x_i / d_i)^{-1}$ $x_i = \text{wt. fraction}$

and any remaining dust is removed in a knockout drum and filter. The cleaned gas can then be sampled and analyzed before the excess is sent to a propane-fired incinerator located on the roof of the building.

3. Gasifier Operating Procedures

Before the start of the gasification experiment the gas chromatograph is checked to ensure it is operating properly and if necessary it is recalibrated. The various bottled gases, including propane, pure oxygen and quenching nitrogen, are then connected to their respective piping. Air from the building air compressor is fed to the inlet of the reactor, the coal feed line and the char recycle line to prevent them from being blocked with bed material. Depending on the experimental conditions, the bed material may consist of char from a previous run, or silica sand. The initial static bed depth of a typical run is about 0.45 m and the bed material is charged from the top of the reactor. The reactor is then heated to 500°C using the propane fired preheater. The heating time varies from 1 hour for the char bed to 2 hours for the silica sand. The bed is spouted using air before the coal is blown into the reactor using pure (undiluted) oxygen. As the reactor temperature increases the propane to the propane preheater is decreased gradually and finally shut off when the bed temperature reaches 750°C. The spouting air is gradually replaced with saturated steam. At 0.5 hr after the introduction of coal, if desired, the char recycle system is turned on and the char is conveyed using air to the reactor. The reactor temperature is then controlled at the desired value by adjusting the steam and oxygen flow rates.

One hour after the start of the experiment, gas samples are taken at 20 minute intervals and analyzed for H_2 , CO_2 , O_2 , N_2 , CH_4 , and CO , using a

gas chromatograph. The temperatures, pressures, and various flowrates of the gasifier system are also recorded at the same times as the gas is sampled for chromatographic analysis. A typical experiment lasts from three to five hours to ensure that a steady state gas composition is achieved. The coal feed, pure oxygen, char recycle feed, air and steam are shut off at the desired time. To prevent any fire hazard the ball valves for the coal feed and char recycle lines are turned off (see Figure 1). The reactor is purged with a short burst of nitrogen to remove any trace of oxygen from the reactor. All electrical power to the gasifier is turned off and nitrogen, pure oxygen and propane are shut off at their respective sources. The reactor is allowed to cool overnight.

The next day, the char collected by the cyclones is discharged from the bins and weighed. A sample is analyzed for carbon and ash content. The coal feeding system is cleaned and the weight of coal used during the experiment is determined. The dust filter is dismantled and cleaned for the next run. The reactor cone is lowered and the bed material is weighed. The carbon and ash content for the bed sample is also determined by a combustion test. Then the reactor cone is placed back into position for the next run. If the pressure drop during the experiment is excessive, (>124 kPa) then the piping in the system is checked and any blockages are cleaned in preparation for the next run. The data collected are processed using the computer program and the results are analyzed. Particle size distributions of the solids found in the bed, and the cyclone catch are determined by sieving. In certain cases the sludge from the dust filter is analyzed for ash, and carbon and its size distribution determined.

The particle size distribution of the crushed anthracite feed is given in Table 3. The Sauter mean diameter, d_p , was 0.83 mm, and the weight mean diameter 1.62 mm. The difference between these values indicates the large amount of fine particles present.

4. Experimental Plans

The overall scheme of the anthracite study is summarized in Table 4. Experiments were run with oxygen-steam blast and no char recycle from the primary cyclone, with air-steam blast, with extra-high temperatures to promote ash agglomeration, with different superficial velocities and, particle sizes, with a K_2CO_3 catalyst, and with a low-ash anthracite in order to accomplish the major experimental tasks of Appendix A.

5. Results and Discussion

5.1 Heating Values, Carbon Conversion and Other Performance Indicators

Before discussing results in detail, it is useful to review what is known about interpretation of results where oxygen, or an air-oxygen mixture or air alone is the oxidant along with steam in the gasification process. Where oxygen-steam blasts are used, to control the temperature below ash fusion conditions high steam/oxygen ratios are used. In this work these are typically 2.4 kg/kg or 4.2 mol/steam/mol. oxygen. Where air is used, values are more typically 0.3 to 0.4 kg/kg or 0.6 mol/mol. This decrease in steam/oxygen mol ratio will generally result in a lowering of the H_2/CO ratio in the gas, given the water-gas shift equilibrium. Simply changing the ratio of H_2/CO in a gas will not affect the heating value since the heat of combustion of H_2 and CO are essentially equal on a molar basis. Secondly, it has been demonstrated in prior work (2), that in gasifying

Table 4: Scheme of Anthracite Gasification Study

Run	Blast	Char-Recycle	Special Conditions
139	O ₂ -Steam	No	
140	"	No	
141	"	Yes	
142	"	Yes	
143	"	Yes	High temperature-Ash agglomeration attempt
144	"	No	High temperature-Ash agglomeration attempt
145	"	Yes	High temperature-Ash agglomeration attempt
146	"	Yes	High temperature-Ash agglomeration attempt
147	"	Yes	
149	O ₂ -Steam	Yes	Low velocity
150	Air-Steam	No	
151	Air-Steam	"	High velocity
152	Air-Steam	"	High velocity
158	O ₂ -Steam	"	Large particle size-catalyst added
159	Air-Steam	"	Large particle size
160	Air-Steam	"	Large particle size
161	O ₂ -Steam	No	Low ash coal run

Table 5: Summary of Results for Anthracite Gasification

RUN NO.	BED WT. (kg)	FEEDRATE (kg/h)				M _{H₂} /M _C	M _{H₂} /M _C	TEMP (°C)		G.H.V. MJ/m ³	G.V. m ³ /kg	% C CHARGES -100	DRY GAS ANALYSIS (SV)				CYCLONE CATCH		% H ₂ O DECOMP.	BED %C	RUN DURATION h:min		
		M _g	M _C	M _{H₂}	M _{CH₄}			AVG	MAX				H ₂	CO	CO ₂	CH ₄	N ₂	kg/h				% C	
139	28.4	21.1	24.0	0.0	0.88	0.0	1.41	1.50	988	980	7.63	1.21	45.6	31.8	27.3	39.2	1.7	0.39	22.06	62.7	15.5	63.2	3:50
140	30.0	20.0	19.1	0.0	1.05	0.0	1.99	1.99	913	925	7.76	1.57	58.5	32.3	28.9	3.9	0.87	0.72	15.71	60.2	17.2	47.3	4:00
141	27.3	20.7	22.6	7.3	1.32	1.66	2.32	1.99	972	990	5.40	2.36	76.6	20.8	22.5	40.5	0.46	15.5	8.85	63.2	12.2	45.9	3:00
142	27.4	23.7	17.0	11.6	1.39	1.63	2.60	1.99	932	939	4.43	2.62	70.5	17.4	19.2	38.4	0.34	25.7	10.78	57.7	9.7	31.6	3:30
143	28.4	26.1	12.1	12.0	2.45	1.63	3.58	1.79	1008	1010	4.61	2.52	79.5	20.2	19.0	36.8	0.32	23.7	6.61	54.4	10.1	55.6	3:50
144	35.9	45.7	22.4	13.0	4.5	53.4	0.0	1.72	926	1001	4.20	1.84	69.7	18.3	15.1	51.8	0.52	14.3	7.37	62.2	5.9	2.2	4:00
145	28.4	30.6	25.6	16.4	11.8	49.7	16.3	1.56	1020	1097	5.13	2.39	77.2	23.8	23.2	34.6	0.41	18.0	9.16	60.5	15.4	58.3	3:28
146	28.3	29.4	25.5	12.0	11.8	40.5	16.3	2.13	994	1026	4.16	2.52	82.2	18.4	18.3	35.7	0.35	23.4	5.49	59.1	9.4	36.5	4:00
147	30.4	36.9	24.4	16.2	12.0	43.0	16.3	1.50	906	925	4.59	1.75	56.0	17.3	19.2	37.6	0.46	25.4	12.44	60.5	7.8	35.5	2:37
149	14.1	32.9	21.4	13.4	5.7	30.4	16.3	1.60	980	995	6.20	1.82	63.5	24.1	25.8	36.3	0.44	13.3	9.89	60.2	13.0	45.1	4:05
150	27.1	22.3	20.3	7.4	0.9	0.0	0.33	1.37	970	972	2.55	3.12	47.0	8.3	12.2	15.0	0.20	64.3	18.52	60.7	32.5	54.9	3:51
151	29.3	19.5	48.8	15.1	16.7	6.6	0.0	3.23	859	874	0.0	7.36	66.1	0.0	0.0	10.9	0.0	79.1	10.80	52.9	0	14.0	3:10
152	19.5	30.5	21.0	22.8	102.0	35.3	0.0	1.36	964	971	0.86	5.04	63.2	4.5	5.0	17.4	0.14	72.9	13.51	64.3	8.0	23.8	2:02
159	30.0	66.9	25.7	17.2	12.9	44.6	0.0	2.58	968	1013	5.40	3.63	75.6	30.5	21.8	31.1	0.41	16.2	5.30	67.0	27.3	45.9	3:45
159	21.3	26.5	21.6	19.9	71.2	9.1	0.0	1.49	943	1018	2.0	4.88	70.0	7.2	9.4	15.5	0.2	67.9	9.44	70.0	34.5	46.8	2:33
160	27.1	34.4	23.6	14.5	77.0	27.7	0.0	1.49	814	960	2.2	4.25	76.5	9.0	9.5	16.9	0.0	64.6	8.65	69.3	12.2	54.6	3:05
161	23.2	42.0	25.5	23.2	11.9	41.9	0.0	1.81	949	970	5.87	2.13	64.6	22.4	24.5	31.7	0.5	20.1	13.7	78.2	17.2	74.2	3:00

* MAF BASIS

under air-oxygen mixtures with steam, that the effect of the presence of the nitrogen in the air can be simply corrected for, i.e. at the same temperature, and O_2 /coal ratio, the gas heating value produced in gasifying in oxygen and steam will be the same as that produced in air and steam if the heating value for the latter is expressed on a N_2 -free basis.

This latter fact is useful in that it permits comparisons among runs with oxygen and with air blasts. Also it permits comparison of runs with and without cyclone char recycle since in our tests the cyclone product is conveyed with air back into the bed. This results in a dry gas with about 14-25% N_2 content, versus the no-recycle case where $N_2 \approx 0.4\%$. However, by expressing gas quality on an N_2 -free basis we can make direct comparisons of heating values. These comparisons are valid since on the large scale, steam could be used to convey the cyclone char to the reactor rather than air, and the gas heating value not affected. Thus some comparisons in this report will be made on an N_2 -free basis.

Performance indicators other than gas heating values are described below. Carbon conversion expresses as a percentage the mols of CO , CH_4 and CO_2 in the product gas per mol of carbon in the feed anthracite. Gas yield (G.Y.) represents the total dry gas flow in m^3 at $15.5^\circ C$ and 1 atm. divided by the M.A.F. anthracite feed rate. The efficiency factor is the product of the cold gas efficiency, η , and the M.A.F. heating value of the anthracite fed

$$\eta \cdot \Delta H_{\text{coal}} = G.H.V. \cdot G.Y.$$

The steam decomposition is expressed either as a percentage of steam fed, or as the mass ratio of kg. steam decomposed per kg. M.A.F. coal fed.

5.2 Typical Experimental Result

Figure 2 shows results of a typical experiment. Over four-hours of anthracite feeding, the average bed temperature climbs from about 940°C to 970°C over the first 1.6 hours, and then increases more slowly over a further 30°C over the next 2.4 hours or about 1.3%/h. The gas heating value oscillates to some extent, but is basically constant after the first 1.5 h. There is a slow increase in H_2 and a slight decrease in N_2 over the course of the run. The average gas composition is about 16% H_2 , 15% CO, 53% CO_2 , 0.5% CH_4 and 14% N_2 . The latter is present because of the air used to recycle the cyclone char. Some minor adjustments of oxygen rate were made during the course of the run. The relatively high O_2 /M.A.F. coal ratio which averages 1.74, results in the high CO_2 content of the gas. Carbon conversion is essentially constant over time. Results discussed below use average values over the latter parts of each run.

5.3 Effect of Oxygen/Anthracite Ratio

A series of experiments was carried out to investigate the effect of gasifier temperature. At a given feed and steam rate, the temperature is largely governed by the ratio of oxygen/coal. Figure 3 shows a combined plot of the carbon conversion and temperature versus the oxygen/anthracite feed mass ratio. Data includes runs at anthracite feed rates of 12.1 to 24 kg M.A.F. coal/h. As the oxygen/coal ratio increases, there is a corresponding increase in temperature and carbon conversion. The highest carbon conversions here are about 80%, achieved at O_2 /coal $\sim$ 2.2, and a bed temperature of 1000°C. The increasing O_2 /coal ratio of course provides more oxidizing conditions and drives down the gas heating value. Figure 4 shows this expected trend. The heating value maximum is achieved at low O_2 /coal

**Table 6: Comparative Data at Highest Gas Heating Value for
Anthracite Versus Other Coals**

Coal	Forestburg		Balmer		Suncor Coke		Anthracite
Run	97C	98	103	105	113	115	140
M _C (kg/h)	29.7	34.4	32.5	24.6	37.2	16.1	19.1
M _{O₂} (kg/h)	27.2	27.2	26.4	25.4	21.5	19.6	20.0
M _S (kg/h)	35.5	36.5	39.4	39.6	33.0	31.3	38.1
M _{Ch} (kg/h)	-	-	-	-	-	-	-
M _{O₂} /M _C -	0.92	0.78	0.811	1.03	0.58	1.22	1.05
T _b (°C)	858	864	898	984	919	970	913
G.H.V. (MJ/m ³)	9.23	9.14	9.48	9.16	8.39	8.10	7.76
%C conversion	42.4	71.6	54.7	64.4	46.8	60.4	58.6
Gas Composition (Dry Basis)							
%H ₂	37.8	36.8	37.0	36.8	36.9	33.5	32.3
%CO	30.5	27.4	29.8	31.9	17.0	24.2	28.9
%CO ₂	28.7	32.0	29.6	28.8	40.1	38.2	37.0
%CH ₄	2.6	3.7	3.7	2.3	4.2	1.3	0.9
%N ₂	0.5	0.14	0.0	0.3	0.72	0.7	0.7

values of about 0.9 - 1.0, where heating values of 7.8 MJ/m^3 (209 Btu/ft^3) are reached. These values are not particularly high compared to that achieved with other coals - previous works on sub-bituminous and bituminous coals yielded values of around 9.1 to 9.5 MJ/m^3 as shown in Table 6. Comparison of Figure 4 and Figure 3 indicates that the higher heating value runs are associated with low carbon conversions, and temperatures of 890°C . Clearly in order to improve performance, temperature must be raised, while maintaining low O_2/coal ratios. For the present reactor with its relatively large heat loss, this means preheating of steam or oxygen to permit higher temperatures to be achieved without increasing the oxygen/coal ratio. This could be done by relatively minor modifications to the test apparatus.

The relatively high O_2/coal ratios required to achieve improved carbon conversion, result in increasing CO_2 levels in the gas. Thus on an N_2 -free basis Figure 5 compares runs at different O_2/coal ratios for gas species concentrations. As the oxygen/feed anthracite ratio increases, an almost linear increase in CO_2 occurs, with the CO_2 level increasing from about 35% at $M_{\text{O}_2}/M_{\text{C}} = 1.05$ to 49% at $M_{\text{O}_2}/M_{\text{C}} = 2.15$. The runs shown in this have a temperature of $\approx 995^\circ\text{C} \pm 14^\circ\text{C}$. With the increase in CO_2 , there is a corresponding linear decrease in both H_2 and CO levels caused by the increase in $M_{\text{O}_2}/M_{\text{C}}$.

Figure 6 shows the efficiency factor

$$\eta \cdot \Delta H_{\text{C}} = \text{G.Y.} \cdot \text{G.H.V.}$$

which is the product of the gas yield and gas heating value, also plotted versus $M_{\text{O}_2}/M_{\text{C}}$. Results for recycle appear to be shifted slightly from those without, due to the lower internal ratio for runs without any char recycle.

Indeed plotting these same results in Figure 7, where the ab-cissa is the ratio $M_{O_2}/(M_C+M_R)$ where M_R is the M.A.F. recycle rate, brings data for recycle into line with that for no recycle. Note that air-steam blowing as opposed to oxygen-steam blowing appears to result in an optimum at a higher $M_{O_2}/(M_C+M_R)$ ratio, presumably because of the heat required to bring the accompanying nitrogen to gasification temperature.

5.4 Ash Agglomeration

Attempts were made to raise the internal gasifier temperature to bring the ash close to agglomerating conditions locally within the gasifier. Figure 8 shows the direct plot of carbon conversion and gas heating value versus average bed temperatures. Small spherical agglomerates were not successfully produced. Rather in Runs 145 and 146, local overheating in the agglomerating zone produced massive clinkering with lumps which ultimately reached nearly the diameter of the gasifier. Maximum temperatures of 1097°C and 1026°C are listed for these runs in Table 5. Temperature profiles versus time, and corresponding gas heating values are shown in Figure 9. It appears that controlled ash agglomeration will require closer temperature control than could be achieved in these conditions. In Run 145 parts of the bed are seen to reach temperatures in excess of 1250°C, which are almost certain to turn the ash to a slag. The precise level of temperature observed here are uncertain due to failure of the stainless steel sheathed thermocouples at these temperatures. In Run 146 although correspondingly high temperatures were not reached, some bed locations did exceed 1100°C. It would appear that agglomerating ash conditions are reached in the neighbourhood of 1050 - 1100°C, which is somewhat lower than found with (bituminous) Balmer coal in earlier work (2).

5.5 Catalytic Run

A single run was carried out with the addition of K_2CO_3 as a dry-feed catalyst as required by Task 5. It would have been preferable to have more than one catalytic run from which to draw conclusions, nevertheless the following comparisons between catalytic and non-catalytic gasification are indicated from the data in Table 7. Catalyst was applied at about 10% loading on M.A.F. coal, by co-feeding powdered K_2CO_3 from the auxiliary feed system shown in Figure 1.

Three non-catalytic runs are included in Table 7 for comparison with Run 158 which used the catalyst. Run 145 has a similar O_2 /coal feed ratio, but used recycle and had a higher bed temperature. Runs 144 and 159 have about the same temperature as Run 158, but are higher and lower respectively in oxygen/coal ratio. Taken together the following can be said for the use of 10% K_2CO_3 on anthracite coal

- (i) Catalyst give higher gas heating value
- (ii) Catalyst gives much higher H_2 in the product gas, which reflects the > 50% increase in kg. steam decomposed per kg. coal fed.
- (iii) Catalyst gives H_2/CO mol ratios in the product gas of around 1.4 compared to 0.7 to 1.2 for non-catalytic gasification.
- (iv) Efficiency factors are some 66% higher for the catalytic run.
- (v) Carbon conversion is higher at similar bed temperatures, however, non-catalytic Run 145 had a similar carbon conversion to the catalytic run with a bed temperature some $50^\circ C$ higher and the benefit of char recycle.

Thus the catalytic run stands out in nearly all performance parameters as being superior to the non-catalytic runs. This effect clearly requires confirmation, but is in line with experience elsewhere with anthracite, and with high pressure steam gasification tests done in our laboratories.

Table 7: Comparison of Catalytic and Non-Catalytic Gasification

Run	Catalyst	No Catalyst		
	158	145	144	159
Catalyst Loading (kg K ₂ CO ₃ /kg coal MAF)	0.151	0	0	0
(kg K ₂ CO ₃ /kg coal as rec.)	0.104	0	0	0
Coal Feed Rate (kg MAF/h)	17.11	16.4	13.0	19.9
Oxygen Rate (kg/h)	26.0	25.6	22.4	23.9
Nitrogen Rate (kg/h)	13.4	11.8	4.5	78.7
Steam Rate (kg/h)	43.3	44.7	52.4	9.0
Recycle Rate (kg/h)	0.0	16.27	0.0	0.0
Oxygen/Coal Ratio (kg/kg)	1.52	1.56	1.72	1.20
Bed Temperature (°C)	968	1020	986	967
G.H.V. (N ₂ -free) (MJ/m ³)	6.68	6.62	4.50	6.35
Steam Decomposed (kg H ₂ O/kg MAF coal)	0.69	0.42	0.24	0.16
Gas Composition by Volume (N ₂ -free)				
% H ₂	36.4	29.1	21.3	21.1
% CO	26.1	28.3	17.6	29.7
% CO ₂	37.1	42.3	60.5	48.7
% CH ₄	0.5	0.5	0.6	0.6
Efficiency Factor (G.H.V. * G.Y.)	20.33	12.26	7.33	9.8
H ₂ /CO mol Ratio	1.39	1.03	1.21	0.7
% C conversion	75.6	77.2	69.7	68.2

5.6 Internal Temperature and Composition Profiles

Temperature and composition profiles were measured (Task 7) in the gasifier for Runs 158-160 using the probes shown in Figure 10. The location of the different probe positions is given in Table 8, along with the measured temperatures for Run 160. Figures 11a, 11b and 11c give the radial gas composition profiles at heights of 14, 25 and 45 cm. Figure 12 shows the axial profile along the spout, and three radial positions.

At the height 14 cm, which is within the cone section it can be seen that the hydrogen and CO profiles fall off radially as the wall is approached with a slight increase in CO_2 . Further up the reactor, at 25 cm, which is above the oxygen inlet, the radial oxygen profile drops to zero as the wall is approached, with corresponding increase in H_2 , CO and CO_2 . At a height of 45 cm the oxygen has essentially vanished, and the H_2 increases as the CO decreases going from the spout to the wall. The axial profile along the spout shows the influence of the oxygen inlet pipe which is located 12.5 cm above the cone inlet. Thus the O_2 profile goes through a maximum, while the reducing species go through a minimum. Figure 13 shows the full axial profile taken in the catalytic Run 158, where the hydrogen and carbon monoxide contents are much higher.

5.7 Low Ash Anthracite Run

Run 161 was carried out with a low-ash anthracite which contained 10% ash rather than the 30% ash material used in the majority of runs (Table 1). Table 9 lists comparable runs on the two materials. The heating value, steam conversion, efficiency factor, H_2/CO mol ratio, and carbon conversion results all show that the performance of the low-ash coal at O_2/coal ratio

Table 8: Probe Positions and Gas Temperatures

Probe Position	Radial Distance L (cm)	Height Above Cone Bottom Z (cm)	Temperature [*] °C
1A	9.0	14	800
B	4.5	14	955
C	0.0	14	1025
2A	15.0	25	730
B	11.0	25	905
C	6.0	25	550
D	0.0	25	265
4A	15.0	45	739
B	11.0	45	923
C	6.0	45	933
D	0.0	45	1062

* Run 160

Table 9: Gasification of Low Ash and High Ash Coals

Ash Content (% As Received)	10.3	30.1	25.4
Run No.	161	140	159
Coal Feed Rate (kg. M.A.F./h)	23.2	19.1	19.9
Oxygen Rate (kg/h)	25.5	20.0	23.9
Nitrogen Rate (kg/h)	11.9	0.0	78.7
Steam Rate (kg/h)	41.9	38.1	9.01
Recycle Rate (kg/h)	0.0	0.0	0.0
Oxygen/Coal Ratio (kg/kg)	1.10	1.05	1.20
Bed Temperature (°C)	950	913	967
G.H.V.(N ₂ -Free) (MJ/m ³)	7.42	7.76	6.35
Steam Decomposed (kg/kg)	0.31	0.34	0.16
Gas Composition by Vol. N ₂ -Free			
H ₂	28.3	31.9	21.1
CO	30.8	27.5	29.7
CO ₂	40.3	39.2	48.7
CH ₄	0.64	1.3	0.6
Efficiency Factor (GHV*GY)	12.1	12.2	9.8
H ₂ /CO Mol ratio	0.92	1.16	0.71
%C Conversion	64.6	58.7	68.2

of 1.10 lie where expected between results listed for high-ash coals at O_2 /coal ratios of 1.05 and 1.20. Thus one can conclude that there is no particular effect of ash content on the gasification process in the range 10% ash to 30% ash. Outside this range, it is possible that ash effects may be important, particularly at very low ash contents.

5.8 Gas Composition versus Kinetics-Free Model Predictions

Gas compositions were predicted for all anthracite runs using the equilibrium model of reference (1). The equations of the model are presented in Appendix B. Figure 14 shows the experimental versus predicted H_2/CO mol ratios in the produced gas. Results are basically similar to what was observed for other materials of low reactivity such as oil sand coke (1) and Balmer coal (3). Experimentally the H_2/CO ratios are close to unity except for the run with K_2CO_3 catalyst and one air gasification run, in spite of thermodynamic predictions suggesting ratios up to 2.4 were expected. Clearly the steam-carbon and water gas shift reactions are not in equilibrium in the anthracite gasification. On the other hand Figure 15 shows that the prediction of the gas heating value by the model is relatively much better. However, there are several runs where large deviations occur.

If the temperature used in the model calculations is raised by 100 - 300 K above the experimental values of the H_2/CO ratios match the predictions more closely. This suggests that it is not a cooling effect in analyzing the off-gases which accounts for the lack of agreement between equilibrium and experimental compositions.

6. Conclusions

A study of the gasification of crushed anthracite from Mount Klappan has shown:

- (i) Compared to low rank coals the anthracite is relatively unreactive in gasification. It requires temperatures in excess of 1000°C to obtain good carbon conversion in the spouted bed gasifier.
- (ii) In order to achieve high temperatures, oxygen/coal ratios must be high in the pilot unit with its significant heat losses. This results in high CO_2 content in the product gases. Thus typical gases contain about 32% H_2 , 28% CO , 1% CH_4 and 38% CO_2 , and have heating value of 7.7 MJ/m^3 . The CO_2 level here is about 8-10% above that produced from lower rank coals.
- (iii) Carbon conversion increases with O_2 /coal mass ratio, but above a value of about 1 kg/kg, gas quality declines as the CO_2 level increases, and the H_2 and CO levels decrease.
- (iv) Use of K_2CO_3 co-fed as a solid to the reactor increased the carbon conversion, and raised the H_2/CO ratio, the heating value, the steam decomposition and the efficiency. One catalytic run only was carried out.
- (v) There was no significant difference in gasification behaviour between low ash (10.3% Ash) and high ash (25-30% Ash) anthracites as far as gas quality and yield were concerned, however, only one run was done for the low ash coal.
- (vi) Attempts to provide smooth ash agglomeration with millimeter sized agglomerates were unsuccessful, in that either agglomeration did not occur presumably due to low temperatures, or sufficient overheating occurred that massive agglomeration occurred.
- (vii) The kinetics-free equilibrium mathematical model predicts the dry gas heating value with reasonable accuracy, but yields H_2/CO levels much higher than observed.

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3. A.P. Watkinson, G. Cheng and C.J. Lim, "Oxygen-Steam Gasification of Coals in a Spouted Bed". In Press. Can. J. Chem. Eng. 1987.

ACKNOWLEDGEMENT

Anthracite for the experiments was kindly supplied by Gulf Canada Corporation, Coal Division.

NOMENCLATURE

d_i	diameter of inlet orifice	mm
d_p	volume-surface mean particle diameter	mm
G.H.V.	gas heating value	MJ/m ³
G.Y.	gas yield	m ³ /kg MAF coal
ΔH	heating value of coal	MJ/kg MAF coal
H	static bed height	m
H_m	maximum spoutable bed depth	m
K_1	equilibrium constant	--
K_2	equilibrium constant	(kPa) ⁻²
M_A	air mass flowrate	kg/h
M_C	coal mass flowrate	kg MAF/h
M_O	oxygen mass flowrate	kg/h
M_S	steam mass flowrate	kg/h
n	total molar flowrate of gases	mol/h
$n_{i, F}$	molar flowrate of species i in feed	mol/h
p	total pressure	Pa
T_b	average bed temperature	K
U_{mf}	minimum fluidization velocity at temperature	m/s
U_{ms}	minimum spouting velocity at temperature	m/s
U_o	superficial gas velocity at temperature	m/s
X	fraction of carbon converted	--
y_i	mol fraction of gas of species i	--
η	thermal efficiency	
ρ_p	particle density	kg/m ³

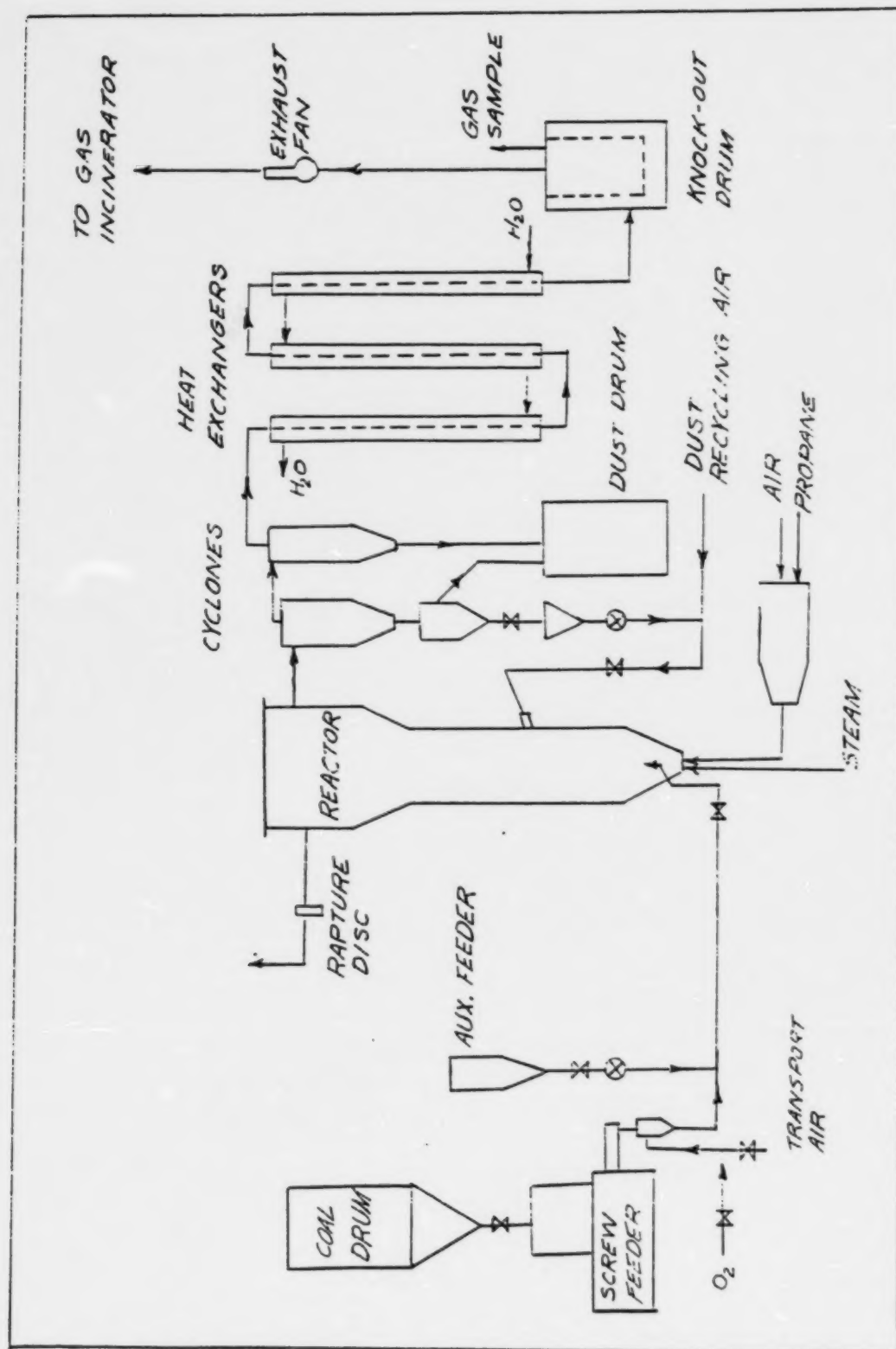


Fig. 1 Schematic Diagram of Apparatus

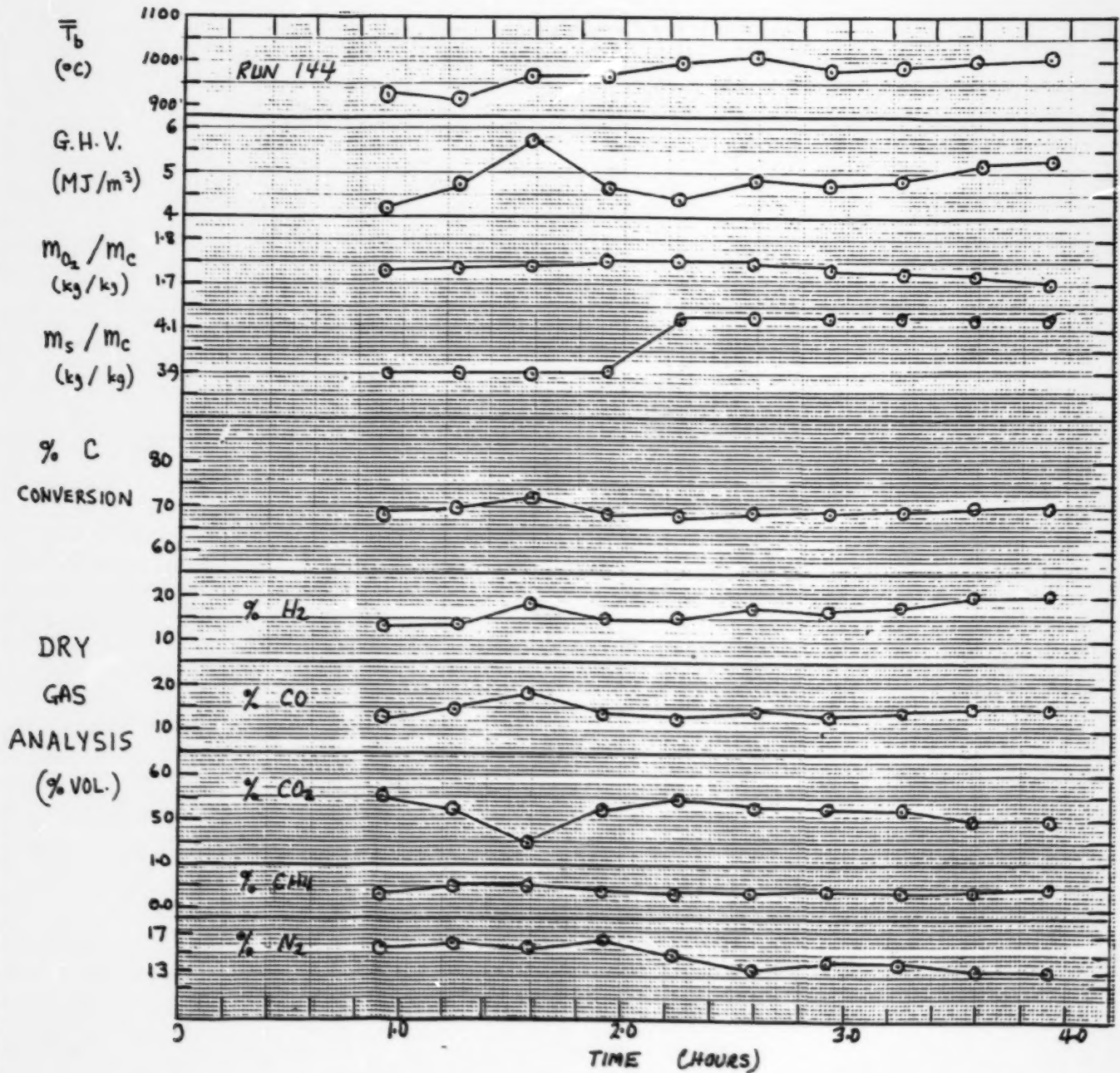
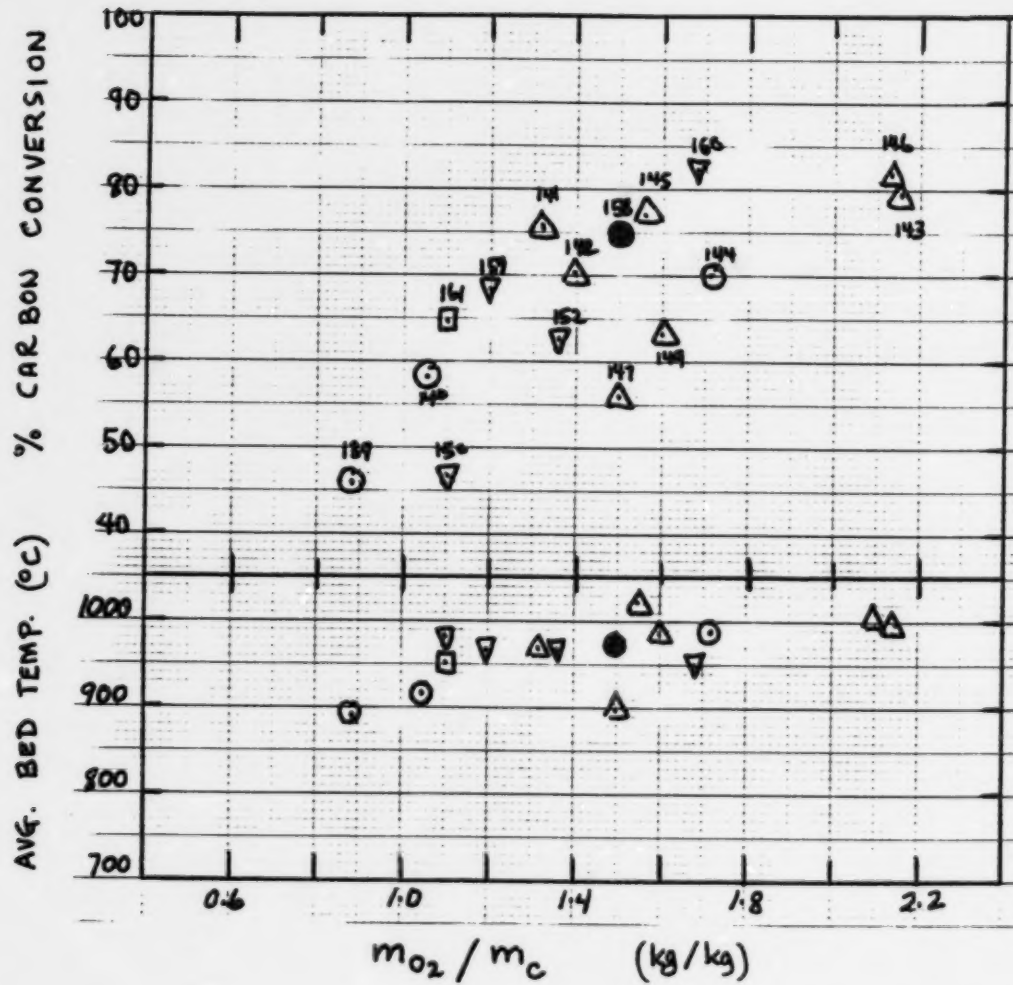


Fig. 2 Result of Typical Experiment



LEGEND

	GAS'N BLAST	RECYCLE	OTHER
○	O ₂ + STEAM	NO	
□	O ₂ + STEAM	NO	LOW ASH COAL
▽	AIR + STEAM	NO	
●	O ₂ + STEAM	NO	CATALYST
◐	O ₂ /AIR + STEAM	YES	

Fig. 3 Carbon Conversion and Temperature Versus Oxygen to Anthracite Feed Rate Mass Ratio

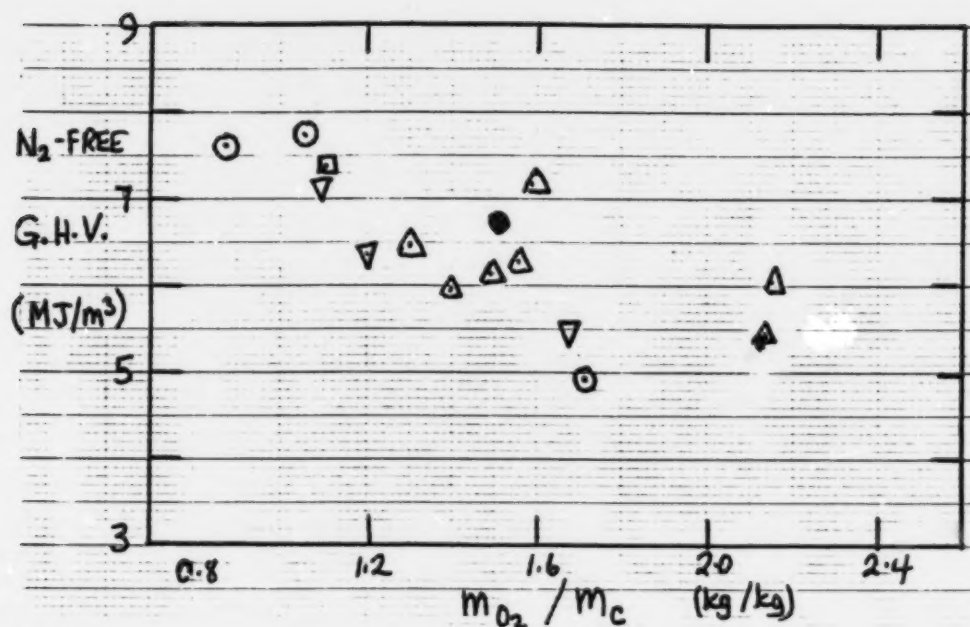


Fig. 4 Gas Heating Value versus Oxygen to Anthracite Feed Rate Mass Ratio

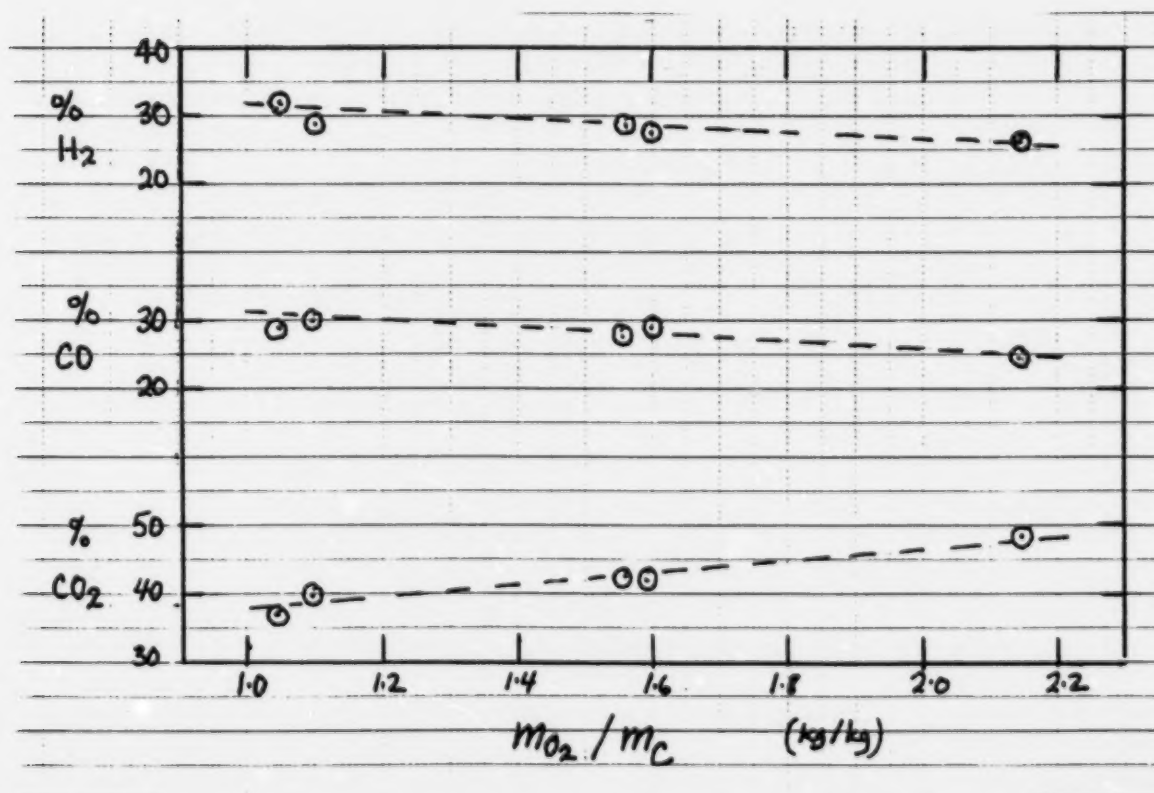


Fig. 5 Gas Composition versus Oxygen to Anthracite Ratio

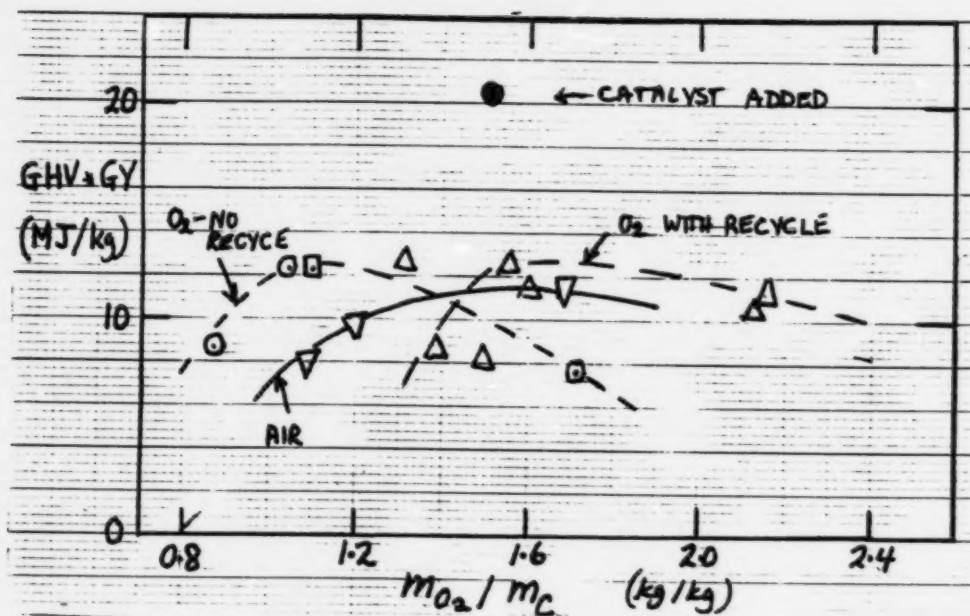


Fig. 6 Efficiency Factor versus Oxygen to Anthracite Ratio

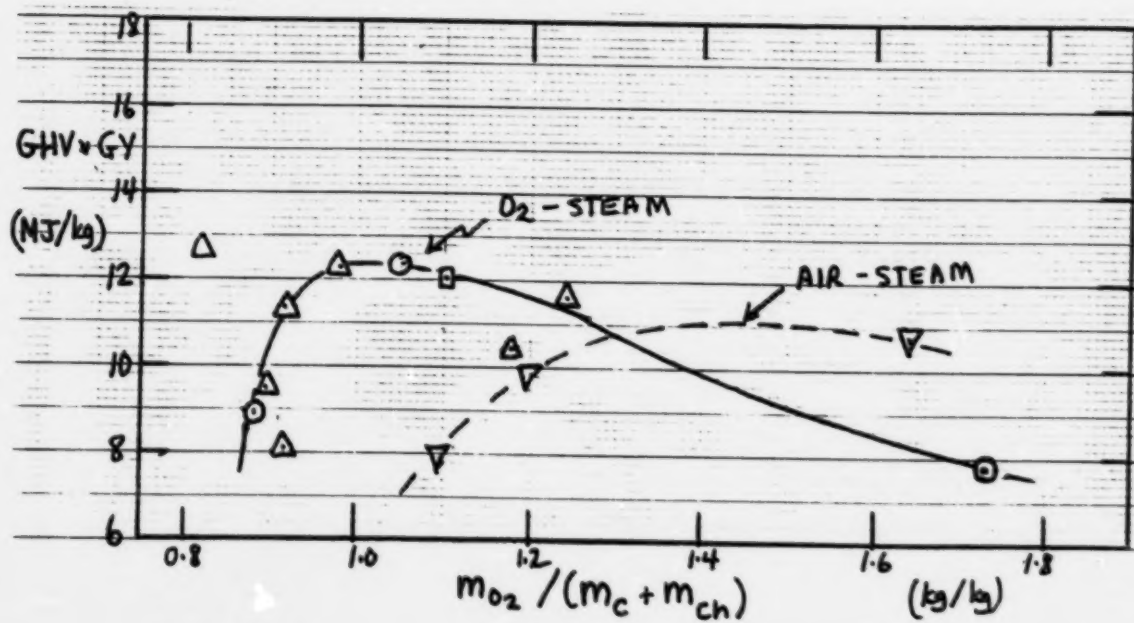


Fig. 7 Efficiency Factor versus Ratio of Oxygen to Feed Plus Recycle Rate

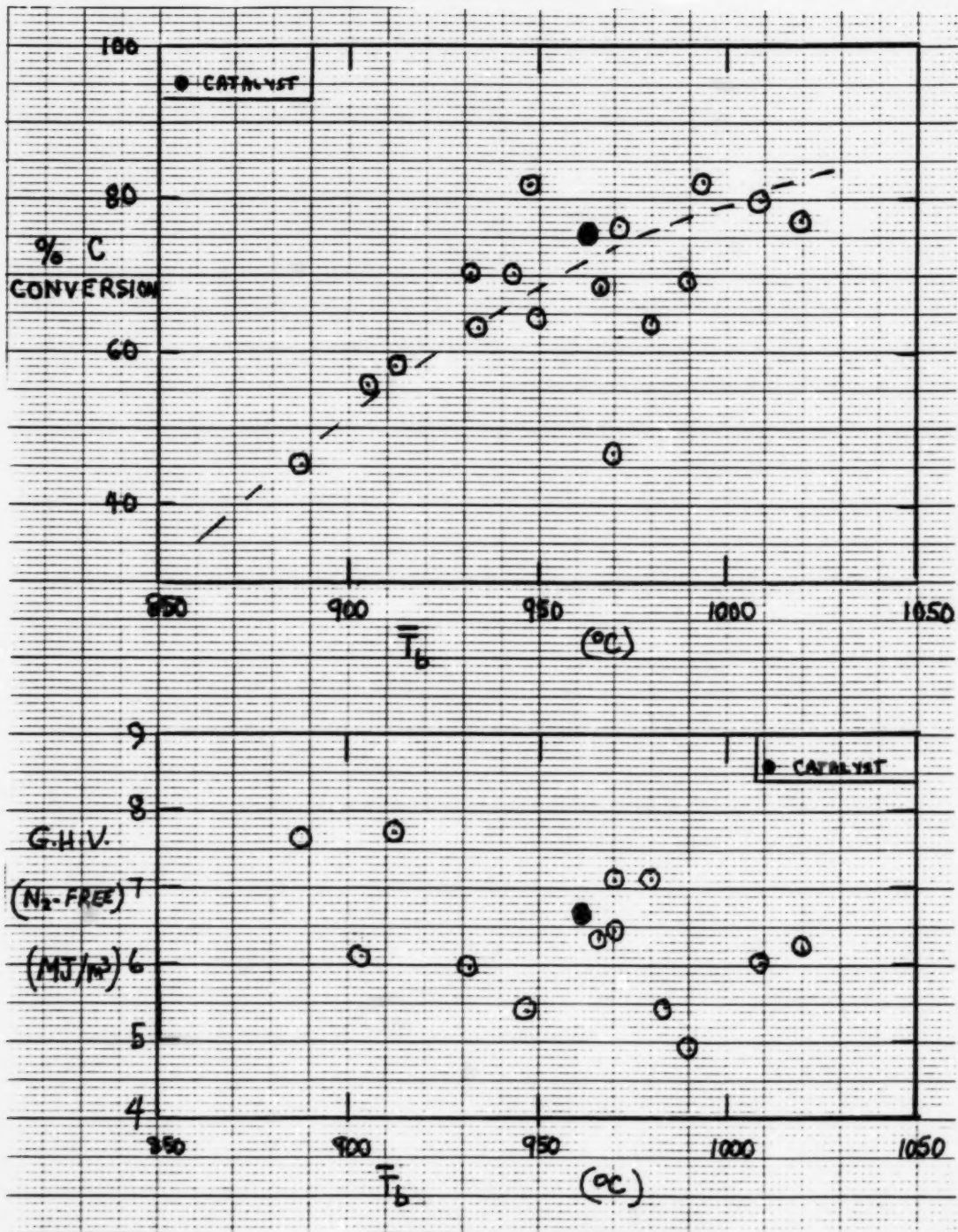


Fig. 8 Carbon Conversion and Gas Heating Value versus Average Bed Temperature

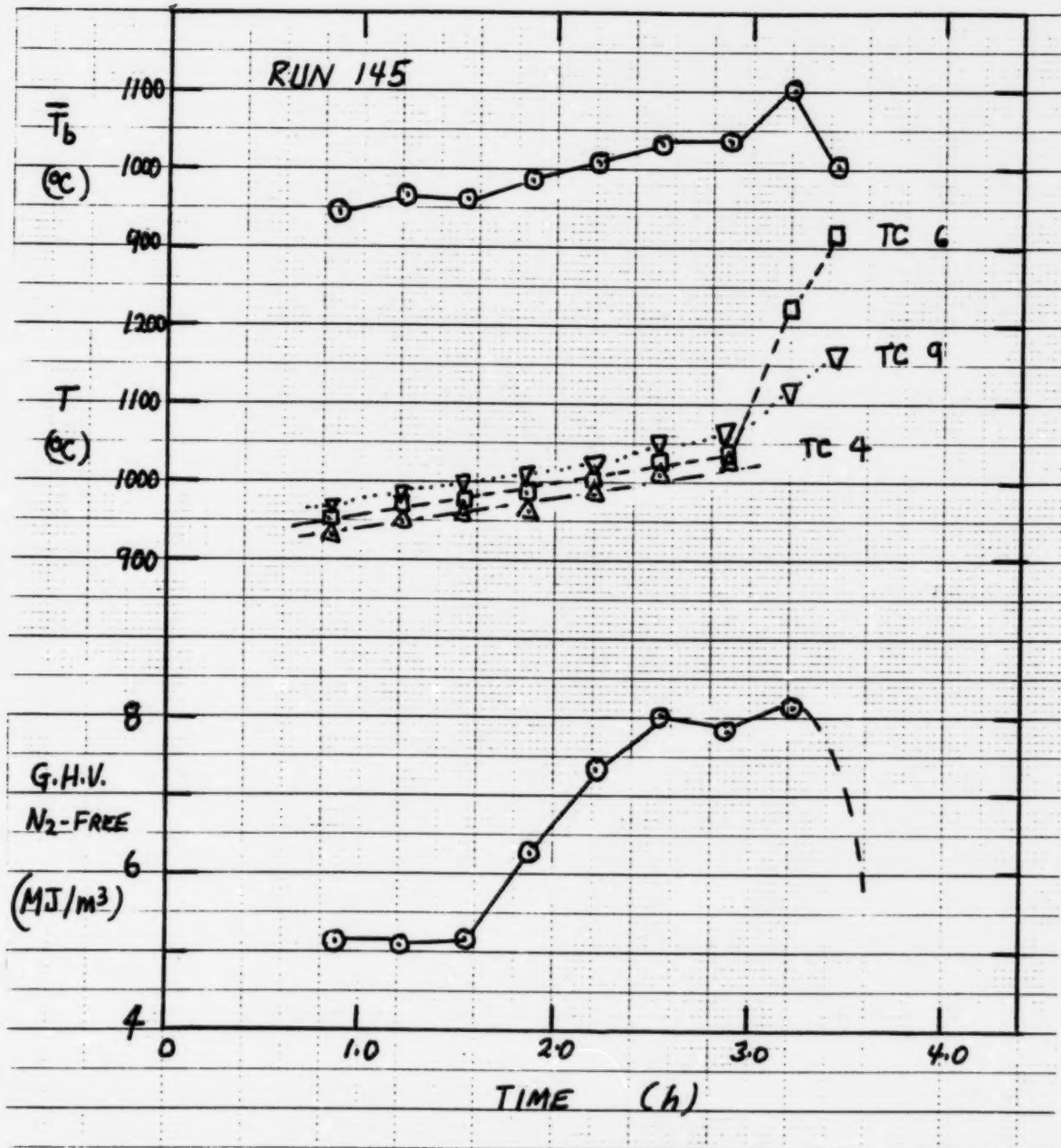


Fig. 9 Typical Temperature Profiles during Ash Agglomeration Attempt

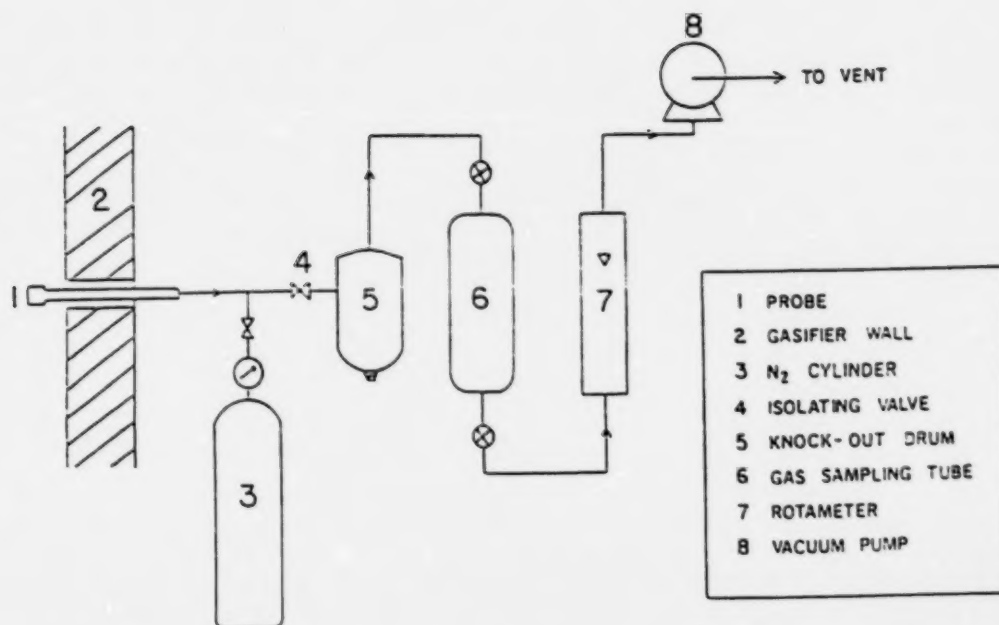
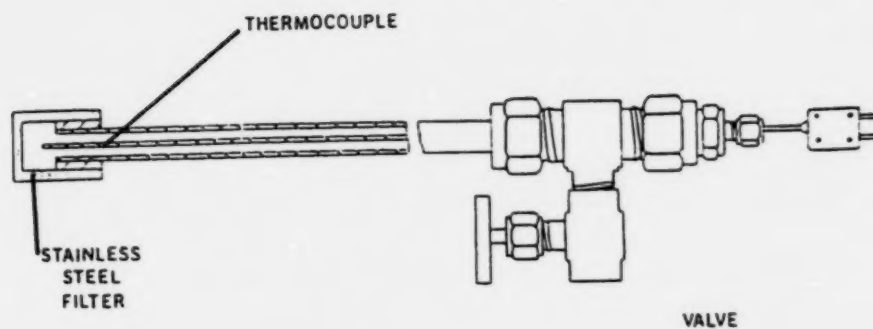


Fig. 10 Internal Gas Composition and Temperature Probe

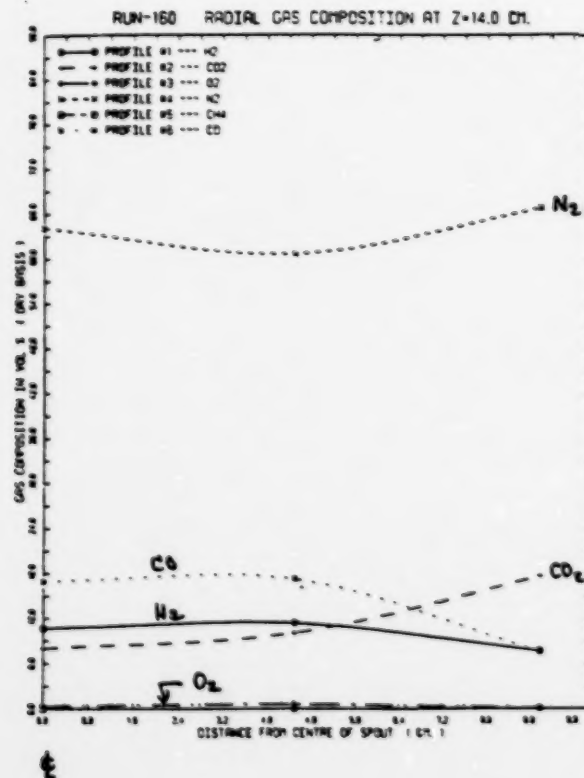
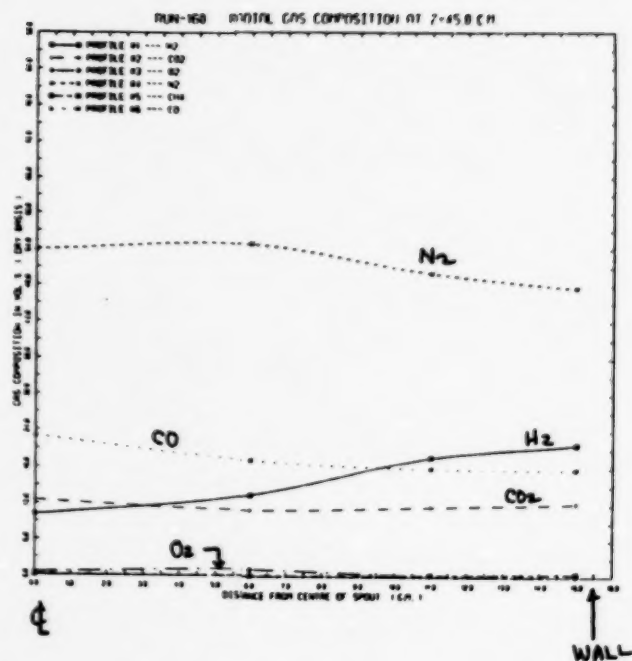
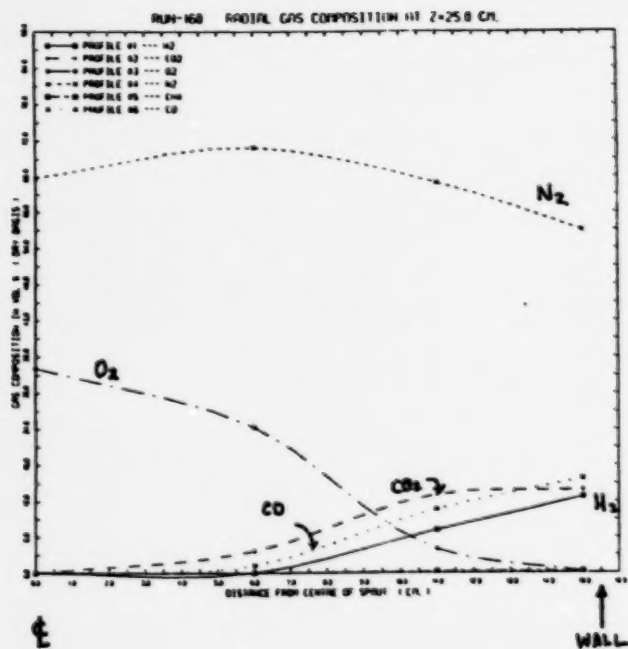


Fig. 11 Radial Gas Composition Profiles at Three Different Axial Positions - Run 160

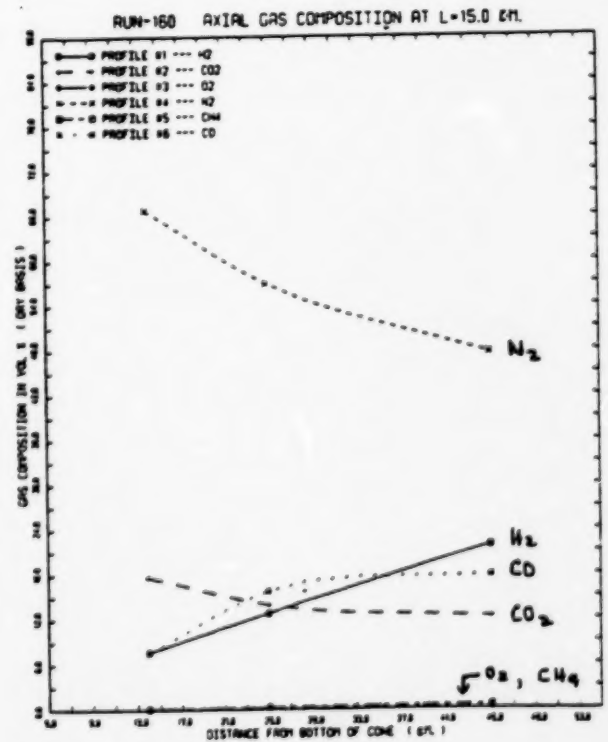
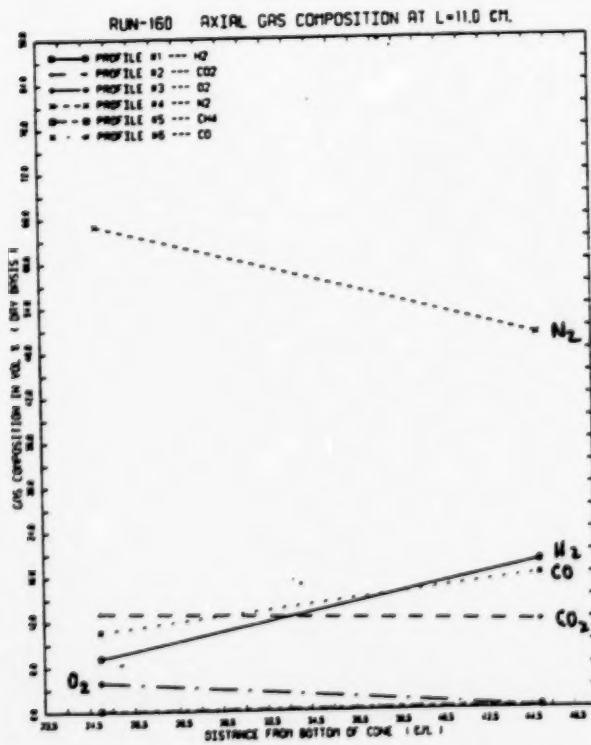
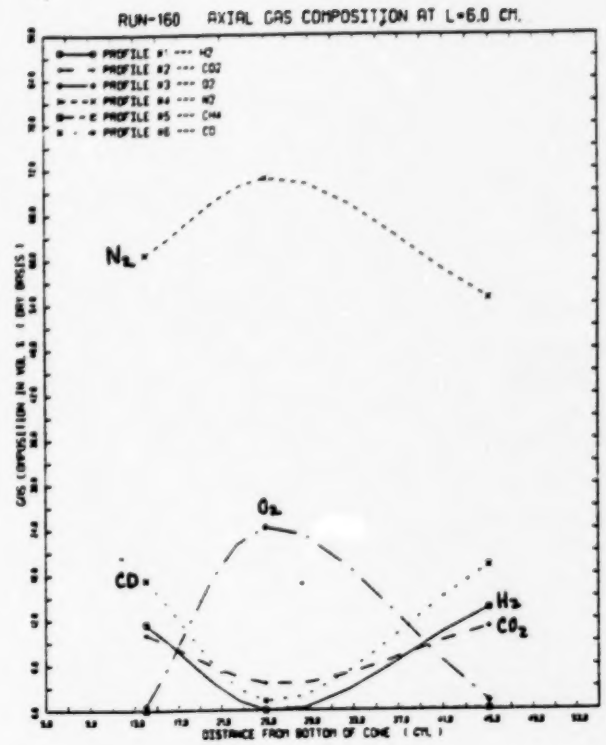
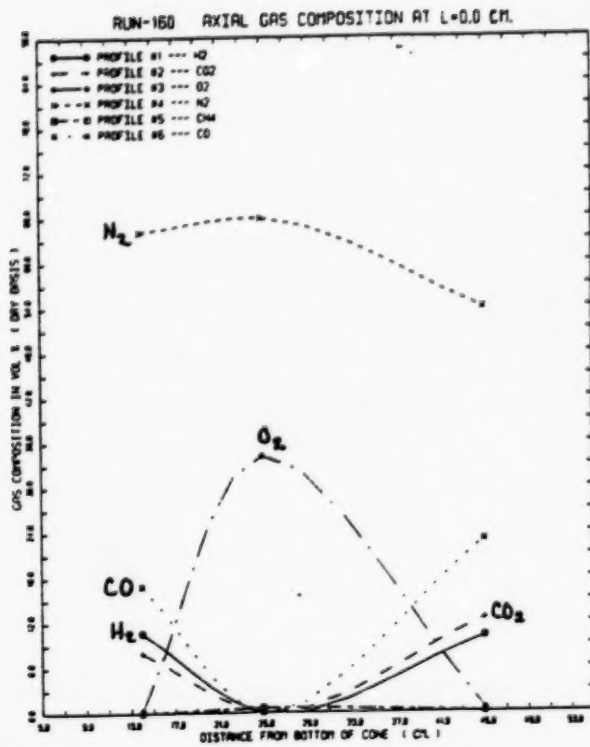


Fig. 12 Axial Gas Composition Profiles - Run 160

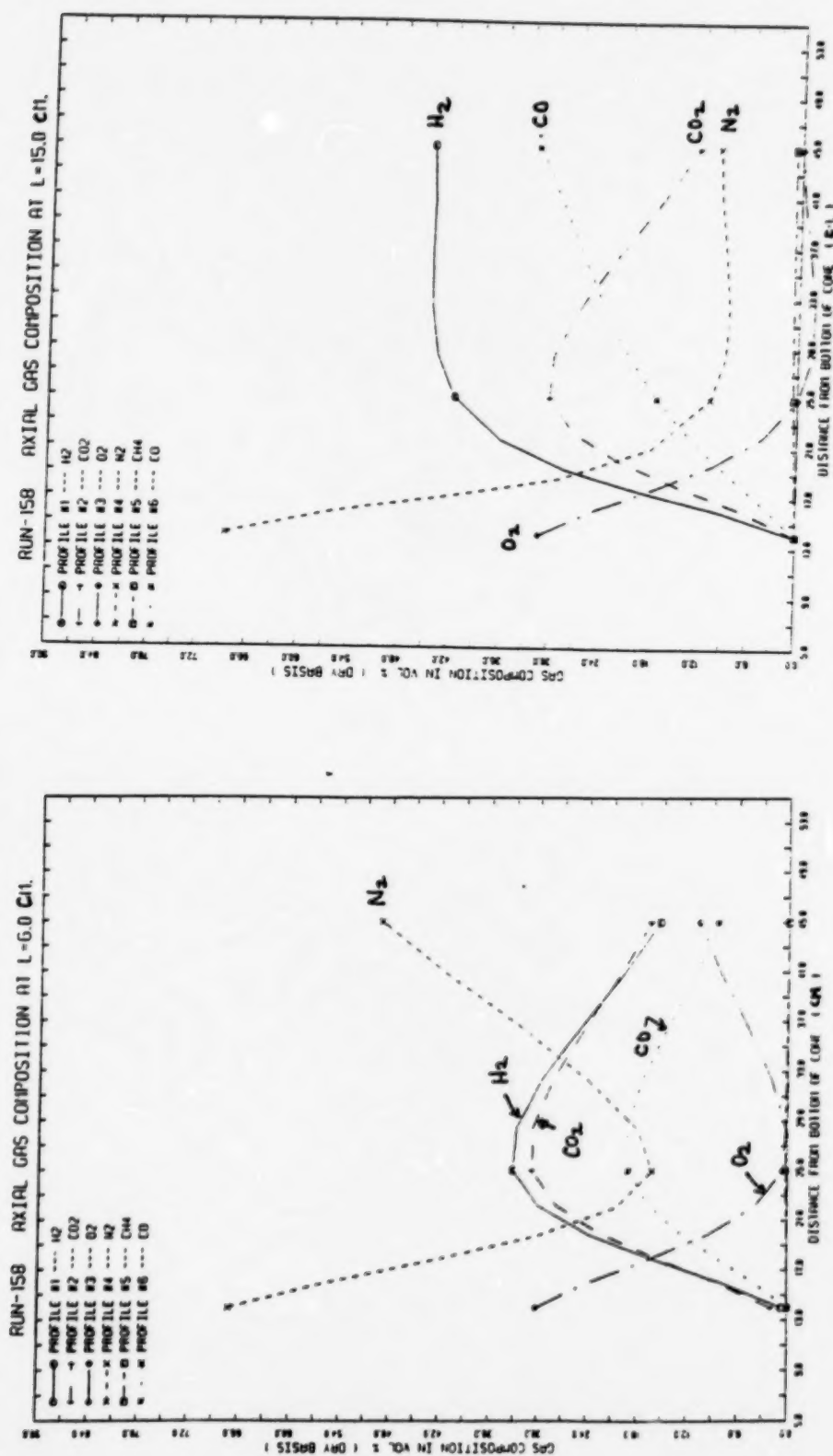


Fig. 13 Axial Gas Composition Profile - Run 158

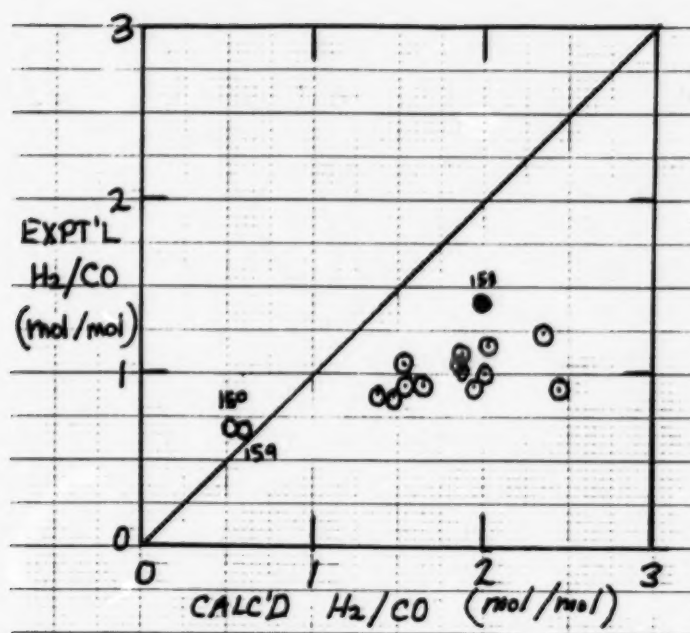


Fig. 14 Experimental versus Predicted H_2/CO Mol Ratio

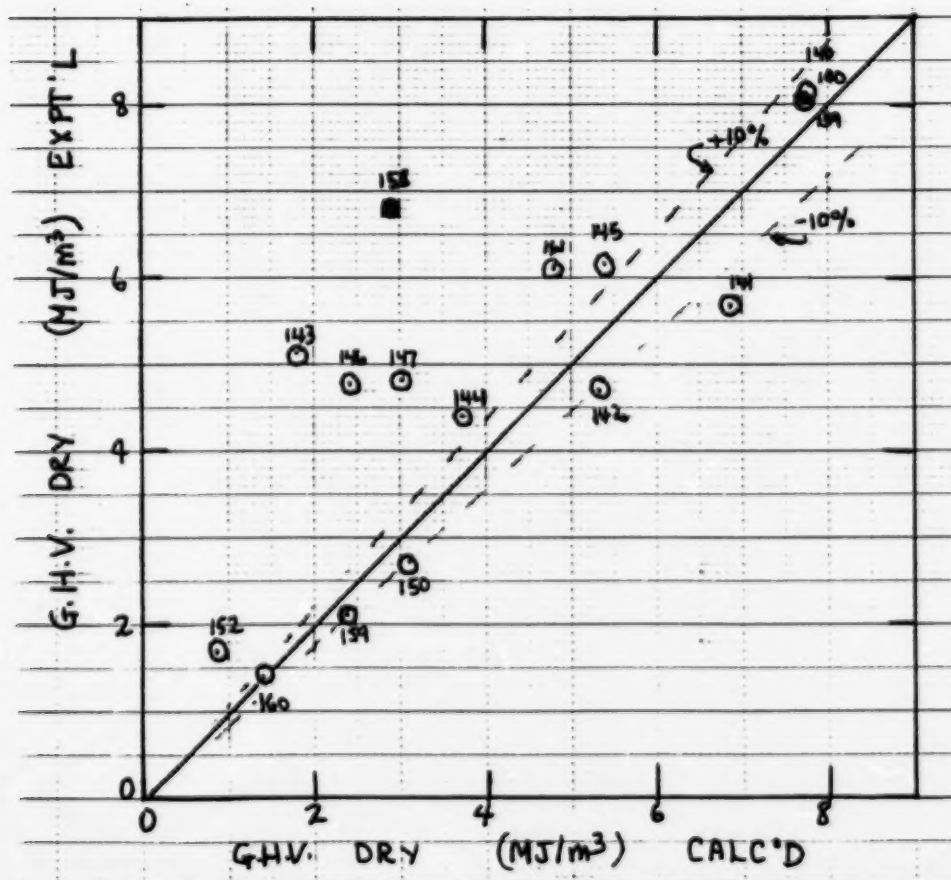


Fig. 15 Experimental versus Predicted Dry Gas Heating Values

APPENDIX "A"
STATEMENT OF WORK

WORK STATEMENT

Title: Gasification of Canadian Anthracite Coal in Fluidized Bed Reactor

Background: Preliminary combustion experiments at the University of British Columbia indicated that the Canadian anthracite coal at the Mount Klappan deposit was quite reactive as compared to some western bituminous coals. It is reasonable to believe that the Mount Klappan anthracite coal can be gasified into gaseous fuel of medium calorific value. The product gas ($H_2 + CO$) can also be used for the synthesis of liquid fuels.

Objectives: (1) To provide an option for utilization of high ash content (20 wt%) Canadian anthracite via the gasification process
(2) To complement the in-house research on the correlation study of coal rank and gasification reactivity

The contractor shall perform the following tasks using the same fluidized bed reactor which was constructed for a previous contract study (DSS No. 12ST.23440-3-9082):

Task 1 Coal Sample Preparation and Minor Modification of Gasifier

About 2 tonnes of Mount Klappan anthracite samples will be supplied for the study by Gulf Canada Limited through the scientific authority. Screening of sample to suitable particle size and minor modification of the reactor configuration may be necessary before commencement of Task 2.

Time required: 0.5 month

Task 2 Gasification of Mount Klappan Anthracite

The gasification characteristics of Mount Klappan anthracite coal will be assessed in a fluidized bed gasifier under various operating modes such as:

- oxygen and steam
- with inert bed
- dry ash
- dust recirculating

The exact operating conditions for the gasification experiments will be selected by Dr. A.P. Watkinson in consultation with the scientific authority. It is preferable to perform the experimental runs at an optimum condition within the range of following parameters:

Sample particle size:	to be determined
Reactor diameter:	0.3 meter
Coal feedrate:	40-50 kg/h
Oxygen/Coal ratio:	0.9 to 1.0 kg/kg
Steam/Oxygen ratio:	to be determined
Minimum fluidizing velocity:	0.4 to 0.7 m/s

Time duration: 2 hours warm up + 5 hours run

A minimum of 12 experimental runs shall be performed. The last two runs shall be reserved for experimental conditions at an optimum level.

The experimental runs can be performed at a rate of one per month to fit into a current contract work schedule.

The carbon balance, carbon conversion, chemical composition of the product gas, dry gas yield, cold clean gas efficiency shall be obtained from each experiment and reported.

Time required: 4 months

APPENDIX "A" (CONT'D)
STATEMENT OF WORK (CONT'D)

Task 3 Chemical Analysis of Char

Proximate and ultimate analyses of the feedstock and selected char samples shall be conducted. These analyses may be subcontracted to outside laboratories.

Time required: 0.5 month + subcontractor cost

Task 4 Compilation of Monthly and Final Reports and Presentation

Brief monthly progress reports of the work will be submitted to the CANMET scientific authority, which would include:

- (i) extent of various tasks completed
- (ii) amount of funds spent

The final report will give a detailed description of the work performed during this project. It should have the following format:

- (i) Introduction;
objectives, justification and usefulness to the Canadian Industry.
- (ii) Experimental;
details of the experimental set-up and description of a typical run.
Details of data evaluation and product characterization.
- (iii) Results and Discussion;
Description of results which were used to arrive at the optimum operating conditions. Comparison of the gasification performance of the Mount Klappan anthracite with other bituminous coals under similar conditions.

All the raw and processed data should be attached in a suitable form (tables, graphs, correlations, etc).
- (iv) Conclusions and Recommendations
Important conclusions related to the gasification of Mount Klappan anthracite should be provided. Specific recommendations for the future course of R & D work in the area should be outlined.

ADD THE FOLLOWING CHANGE:

Task 4 Renamed as Task 8

Additional Tasks:

- Task 4 Clean the coal sample from 32% ash to about 8-10% ash and gasify the clean coal under the best conditions found in the tests at 32% ash.
- Task 5 Perform one gasification run (3 to 5 hours) to check the effect of potassium carbonate catalyst on carbon conversion of Anthracite gasification. Use the best conditions for the gasification based on catalysis of Balmer coal, Suncor coke, and Syncrude coke in a previous contract study.
- Task 6 Perform two runs (3 to 5 hours) in ash agglomerating mode by raising reactor temperature to improve carbon conversion over that operated in a dry ash mode in the existing contract.
- Task 7 Perform experiments to sample reactor internal composition and temperature profiles. Run the UBC gasifier mathematical model (using the batch kinetic data) to compare predictions vs. performance on a low reactivity Anthracite.

Appendix B Kinetics-Free Model

In the kinetics-free model we assume that the exit gas composition of the gasifier is determined solely by stoichiometric and thermodynamic constraints. The following specific assumptions are made in the model development:

- a) All feed oxygen is consumed
- b) Carbon conversion is given (from experimental measurement)
- c) Exit gases from the gasifier contain H_2 , CO , CO_2 , CH_4 , H_2O and N_2 , and are in water-gas shift and methanation equilibrium at the average gasifier temperature.

The model consists of a series of elemental mass balance equations for C, O, H, and N, the equilibrium equations of the water-gas shift and the methanation reactions, and Dalton's law. They are summarized as follows:

- 1) The equilibrium equations

$$\text{For water-gas shift: } K_I = \frac{(y_{H_2})(y_{CO_2})}{(y_{CO})(y_{H_2O})} \quad (3)$$

$$\text{For methanation: } K_{II} = \frac{(y_{CH_4})(y_{H_2O})}{(y_{CO})(y_{H_2})^3 p^2} \quad (4)$$

where the y are the mole fractions and p is the total pressure of the system (1 atm for our case). K_I and K_{II} are the corresponding equilibrium constants given by the following expressions:

$$K_I = 0.0265 \text{ Exp}(3956/T)$$

$$K_{II} = 6.7125 \times 10^{-14} \text{ Exp}(27020/T)$$

where T is the average gasifier temperature.

2) The elemental mass balance equations

$$\text{For H}_2: (2y_{CH_4} + y_{H_2} + y_{H_2O})n = n_{H_2} \quad (5)$$

$$\text{For O}_2: (y_{CO}/2 + y_{CO_2} + y_{H_2O}/2)n = n_{O_2,F} \quad (6)$$

$$\text{For C: } (y_{CH_4} + y_{CO} + y_{CO_2})n = Xn_{C,F} \quad (7)$$

$$\text{For N: } ny_{N_2} = n_{N_2,F} \quad (8)$$

where $n_{H_2,F}$, $n_{O_2,F}$, $n_{C,F}$ and $n_{N_2,F}$ are the number of moles of hydrogen, oxygen, carbon and nitrogen fed to the gasifier, n is the total molar flows in the exit gases, and X is fraction of carbon converted to gases.

3) Dalton's Law

$$y_{CO} + y_{CO_2} + y_{H_2} + y_{H_2O} + y_{N_2} + y_{CH_4} = 1 \quad (9)$$

For a given gasification experiment, $n_{H_2,F}$, $n_{O_2,F}$, $n_{C,F}$, $n_{N_2,F}$, T and X are known. K_I and K_{II} can be calculated from T . The equations (3) to (9) can be solved by a nonlinear equation routine on the UBC computer to yield exit gas compositions, y_{CO} , y_{CO_2} , y_{H_2} , y_{CH_4} , y_{H_2O} , and y_{N_2} , as well as n , the total molar flow. The H_2/CO molar ratio and GCV of the exit gases can be calculated.